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## ***Analysis and simulation of scale-up potentials in Reverse Electro dialysis***

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### **Abstract**

The Reverse Electro dialysis (RED) process has been widely accepted as a viable and promising technology to produce electric energy from salinity difference (salinity gradient power - e.g. using river water/seawater, or seawater and concentrated brines). Recent R&D efforts demonstrated how an appropriate design of the RED unit and a suitable selection of process conditions may crucially enhance the process performance. With this regard, a process simulator was developed and validated with experimental data collected on a lab-scale unit, providing a new modelling tool for process optimisation.

In this work, performed within the REAPower project ([www.reapower.eu](http://www.reapower.eu)), a process simulator previously proposed by the same authors has been modified in order to predict the behaviour of a cross-flow RED unit. The model was then adopted to investigate the influence of the most important variables (i.e. solution properties and stack geometry) on the overall process performance. In particular, the use of different concentrations and flow rates for the feed streams have been considered, as well as different aspect ratios in asymmetric stacks. Moreover, the influence of the scaling-up a RED unit was investigated, starting from a 22x22 cm<sup>2</sup> 100 cell pairs lab-stack, and simulating the performance of larger stacks up to a 44x88 cm<sup>2</sup> 500 cell pairs unit.

Finally, different scenarios are proposed for a prototype-scale RED plant, providing useful indications for the technology scale-up towards 1 kW of power production, relevant to the installation of a real prototype plant in Trapani (Italy) being the final objective of the R&D activities of the REAPower project.

### **Keywords**

Salinity Gradient Power; RED; sea water; brine; process simulator; model.

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# 1 Introduction

Among the novel renewable energy sources investigated nowadays, a growing interest is being gained by Salinity Gradient Power (SGP), i.e. the energy available from the mixing of two aqueous solutions with different salinity. SGP has been estimated to be the second largest marine-based energy source, with a global power of 980 GW potentially available for extraction [1]. In order to suitably exploit such energy source, a number of technologies have been proposed in the literature, based on membranes [2, 3] or capacitive electrodes applications [4, 5]. Among these alternatives, Reverse Electrodialysis (SGP-RE or RED) can be considered as a truly promising way to directly convert SGP into electricity. In fact, this process can become a reliable industrial application within the foreseeable future, as long as the required membranes will be commercially available at competitive costs [1, 6]. The Reverse Electrodialysis process is based on the controlled mixing of two salt solutions by means of Ion Exchange Membranes (IEMs). A RED system (*stack*) is constituted by a pile of Anion and Cation Exchange Membranes (AEMs and CEMs), which are normally separated by netting spacers to create compartments where solutions can flow (Figure 1). During the operation, the system is fed by a concentrated saline solution (*concentrate* or *HIGH solution*) and a dilute one (*diluate* or *LOW solution*) in alternating compartments. The salinity difference provides the driving force for electromotive force generation and for ions transport from concentrate to diluate channels. Ideally, only anions can pass through AEMs and only cations can pass through CEMs. The overall results is an ionic current through the system, which can be eventually converted into an electric current at the two electrodes (located at the ends of the stack), to be finally used by an external load.

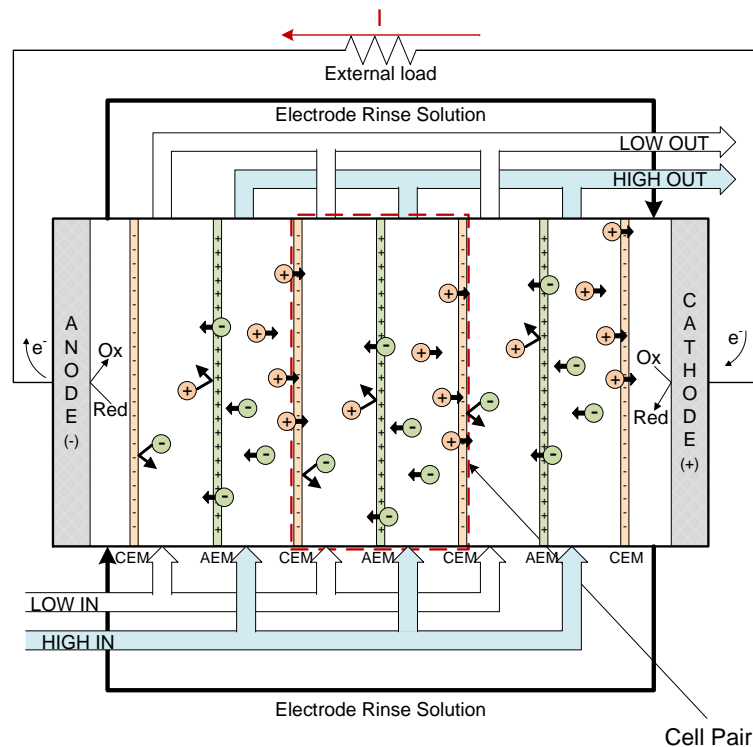


Figure 1. Schematic representation of the RED process.

A number of non-ideal phenomena within the system have been taken into account for a reliable physical description of the process, namely:

- *real IEMs permselectivity*. In real membranes also the co-ions (i.e. ions with the same charge of fixed ions within membranes) will pass to some extent through IEMs, thus causing a reduction of the generated ionic current [7].
- *Concentration polarisation phenomena in the membrane-solution boundary layer*. The actual concentration at the membrane-solution interface can be different than the bulk concentration due to polarisation phenomena [8, 9]. This leads to a decrease of the effective available driving force.
- *Solvent transport through IEMs*. Water molecules can pass through membranes according to two counter-acting phenomena, i.e. osmosis and electro-osmosis.
- *Parasitic currents*. Inlet/outlet manifolds where saline solutions flow act as shortcut paths for ionic currents, thus generating parasitic currents which deplete the net available electromotive force at the stack electrodes [10].

The mathematical description of the RED process has been already addressed at different detailed levels by many authors [11-15, 23-25]. However, the vast majority of such modelling works essentially refer to the use of fresh water as dilute solution and seawater as concentrate. The use of highly concentrated salt solutions can strongly affect the behaviour of both solutions and membranes, thus requiring a different mathematical approach [16].

Focus of this work is to present a newly developed process simulator for the RED process with sea/brackish water and concentrated brines as feed streams. Starting from a model previously developed by the same authors [17], a more detailed simulation tool was built, allowing to quantitatively explore the effect of new stack design, e.g. a cross-flow configuration with asymmetrical paths for the feed solutions.

The developed simulation tool and validation procedure are described in detail in Section 2. Model predictions are reported in Section 3, starting from a comparison with the model previously proposed for a co-current flow configuration [17]. Finally, Section 4 summarises the main results achieved with respect to the simulation of a prototype-scale unit targeting a power output of 1 kW.

## 2 Description of the modelling approach

A Multi-Scale Modelling approach has been applied to describe the system (Figure 2) [18]. According to this method, a lower-scale model describes the physical phenomena in a single repeating unit of the stack (*cell pair*). The second level of description refers to the whole “*stack*”, including all cell pairs and the relevant interconnections. Finally, a “*plant*” model has been implemented, allowing to simulate more stacks connected with different layouts.

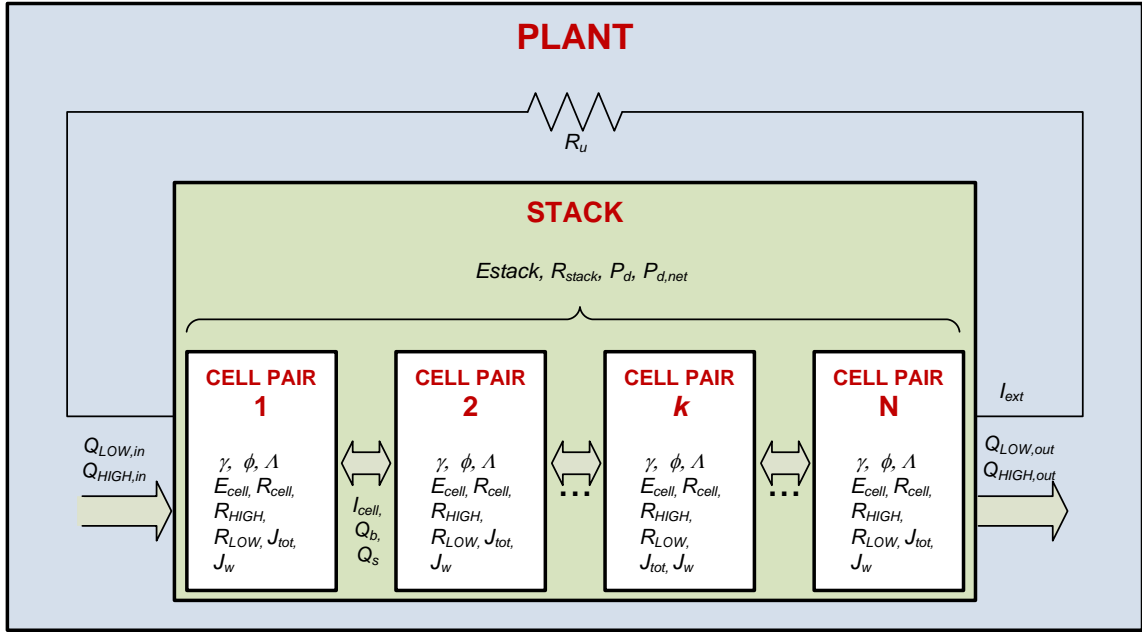


Figure 2. Hierarchical structure of the proposed model.

## 2.1 Model development and implementation

A model for RED process using sea/brackish water as diluate and brines as concentrate streams was already presented by the authors [17], aiming at a comprehensive simulation tool for the RED process accounting for all the above mentioned non-ideal phenomena. The complete set of model equations can be found in [17]. This paragraph illustrates only the new implementations carried out within the present work, in order to simulate the operation of a cross-flow stack with possible asymmetric configurations of the feed channels.

### 2.1.1 Development of a 2-D model structure to simulate cross-flow arrangement

It is worth noting that the model described in [17] was developed for a co-current flow distribution, where all the variables dependent on salt concentration (solution conductivity, cell pair voltage, etc.) vary only along compartment length. For this reason, a 1-D distributed model was considered, where the computational domain was discretised along the compartment length (Figure 3a).

In this work, the previous model was purposely modified in order to account for a different flow arrangement, with cross-flow currents. In this case, the computational domain is defined in 2-Dimensions, where  $x$  and  $y$  are the flow directions for the concentrate and dilute solutions, respectively (Figure 3b).

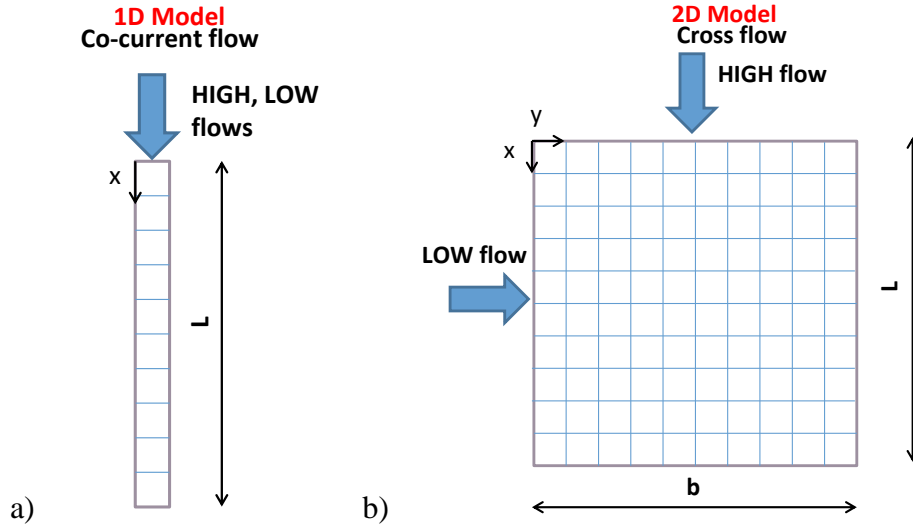


Figure 3. Computational domain of developed models: a) 1-D model [17]; b) 2-D model (this work).

The main advantage of such 2-D model approach is the high flexibility to simulate the use of asymmetric stack geometries, with one compartment longer and the other one wider, thus changing the aspect ratio of the system.

The new computational domain required a number of modifications within the model formulation. In particular, all distributed variables described as  $x$ -dependent in [17] can vary along both  $x$  and  $y$  coordinates. Likewise, the mass balance in diluate and concentrate compartments have been modified as:

$$\frac{\partial(Q_{HIGH} C_{HIGH}(x, y))}{\partial x} = -J_{tot}(x, y) dy ; \quad \frac{\partial(Q_{LOW} C_{LOW}(x, y))}{\partial y} = J_{tot}(x, y) dx \quad (1,2)$$

$$\frac{\partial(Q_{HIGH}(x, y))}{\partial x} = -J_{w,net}(x, y) dy ; \quad \frac{\partial(Q_{LOW}(x, y))}{\partial y} = J_{w,net}(x, y) dx \quad (3,4)$$

where  $Q$  is the volumetric flow rate in each compartment,  $C$  is the molar concentration,  $J_{tot}$  and  $J_{w,net}$  are the salt and solvent fluxes through membranes. Subscripts *HIGH* and *LOW* refer to concentrated and dilute solutions, respectively.

### 2.1.2 Response variables of the process

For the sake of brevity, only the model response variables are reported. In particular, the gross power produced by the system is:

$$P = I_{ext}^2 R_u \quad (5)$$

where  $I_{ext}$  is the electric current in the external circuit, and  $R_u$  is the external load connected to the stack. The required power for pumping the salt solutions through the stack is calculated by:

$$P_{pump} = \frac{\Delta P_{HIGH} Q_{HIGH}^{tot} + \Delta P_{LOW} Q_{LOW}^{tot}}{\eta_p} \quad (6)$$

$\Delta P$  is the total pressure drop,  $Q^{tot}$  is the flow rate of both solutions, and  $\eta_p$  is pump efficiency (assumed to be 75%); subscripts *HIGH* and *LOW* refer to concentrate and diluate, respectively. Pressure drops for both solutions were estimated based on experimentally measured data for a single channel filled with a 270  $\mu\text{m}$  polyamide woven spacer [17].

The power density is defined as the electric power produced by the system per unit cell pair area:

$$P_d = \frac{1}{N} \left( \frac{I_{ext}}{A} \right)^2 R_u \quad (7)$$

where  $N$  is the number of cell pairs within the RED unit, and  $A$  is the membrane active area. Finally, the net power density is given by:

$$P_{d,net} = P_d - \frac{P_{pump}}{N A} \quad (8)$$

In eq. (8) the pumping power ( $P_{pump}$ ) has been divided by the cell pair area ( $N A$ ) to be dimensionally consistent with the power density.

## 2.2 Validation of the model on a lab-scale unit

The newly implemented model was validated by experimental data collected with a 50 cell pairs-stack with cross-flow arrangement. Experimental tests were performed changing the inlet concentration of diluate and concentrate [17]. Results are shown in Figure 4.

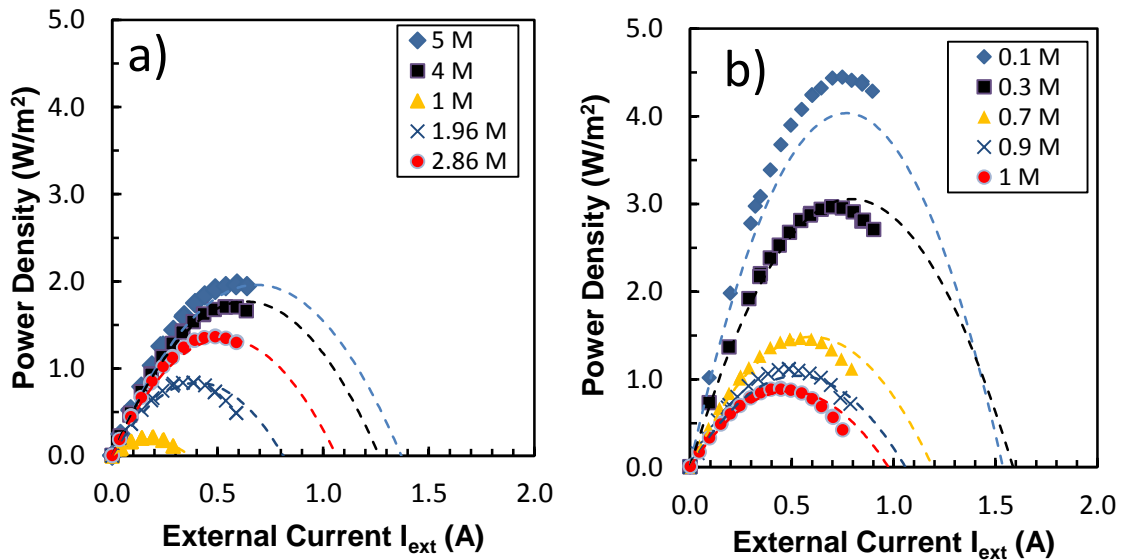


Figure 4. Effect of a) HIGH and b) LOW inlet concentration on Power density. Experimental data (points) and model predictions (curves) for a 50-cells stack equipped with Fujifilm

membranes  $10 \times 10 \text{ cm}^2$  and  $270 \text{ }\mu\text{m}$  woven spacers; a)  $C_{\text{LOW}} = 0.55 \text{ M}$ ; b)  $C_{\text{HIGH}} = 5 \text{ M}$ ; fluid velocity:  $1 \text{ cm/s}$ ;  $T = 20^\circ\text{C}$ .

Figure 4 shows how the model can predict with good agreement the experimental behaviour of the system in a wide range of feed concentrations.

### **3 Simulation results and analysis of perspectives**

#### **3.1 Comparison 1-D and 2-D model**

Unlike the 1-D model, the new one can evaluate all the relevant 2-D distributed variables on membrane surface. As an example, Figure 5 shows the concentration profiles predicted by both models for a  $20 \times 20 \text{ cm}^2$  stack equipped with 50 cell pairs.

Interestingly, the 2-D maps indicate that the main concentration variation occurs along the flow direction, which was the only one predictable by the 1-D model. However, for cross flow stack, this variation occurs for the two solutions in perpendicular directions, as can be predicted only by a 2-D model. Moreover, a less important (yet predictable) variation in concentration occurs also in the direction perpendicular to the flow only within the diluate compartment, while a pretty monodimensional behaviour can be observed for the concentrate one.

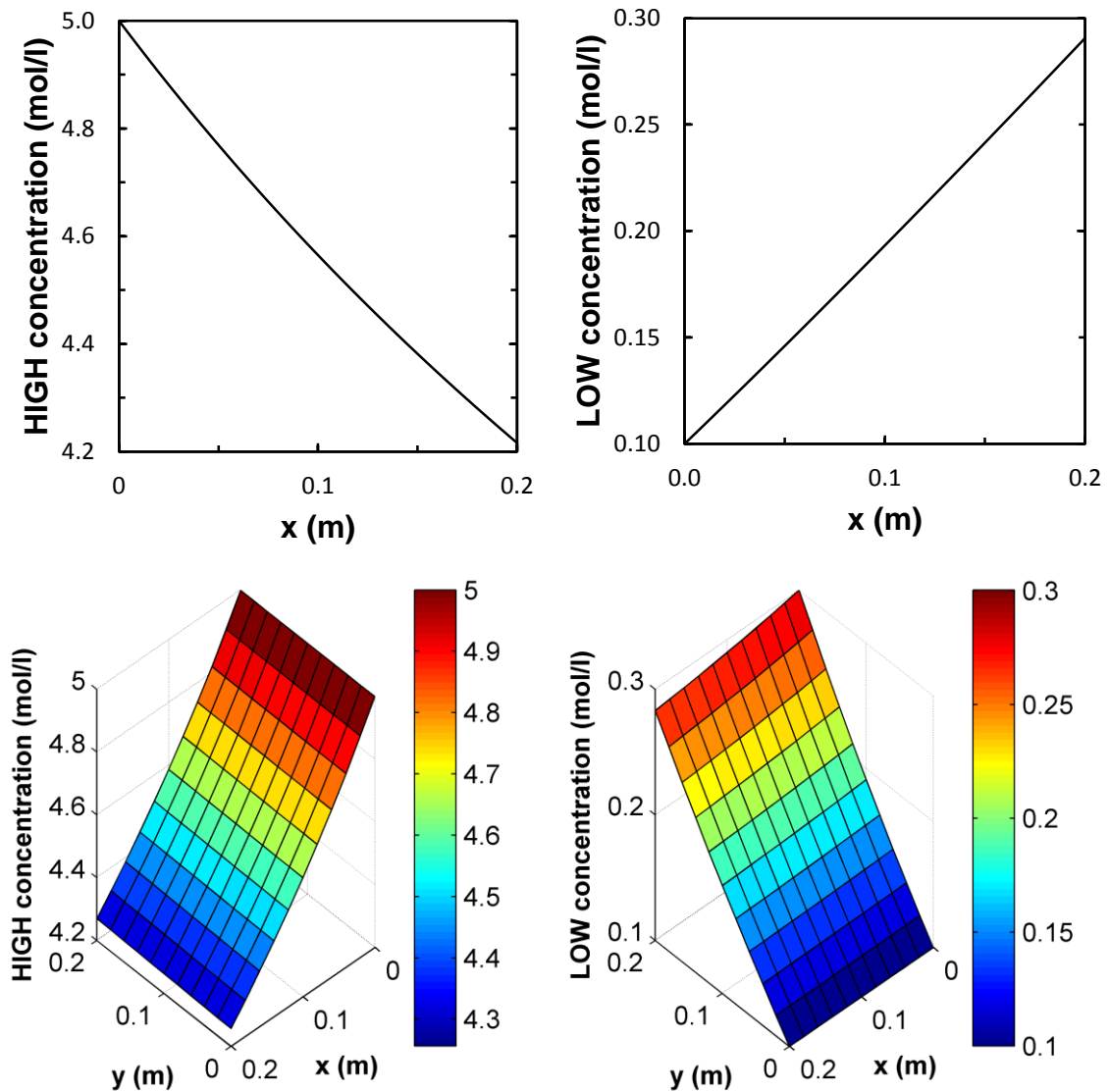


Figure 5. Concentration profiles within compartments estimated by the previous 1-D co-current model (above, [17]) and by the 2-D cross-flow model (below). Simulations of a  $20 \times 20$  cm<sup>2</sup> 50 cell pairs stack equipped with Fujifilm membranes, 270  $\mu$ m woven spacers;  $C_{\text{HIGH}} = 5$  M,  $C_{\text{LOW}} = 0.1$  M; fluid velocity: 0.5 cm/s;  $T=20^\circ\text{C}$ .

As a result of the predicted change in concentrations in the two “cross-flow” channels, the variation of cell voltage along the two directions is also predicted by the 2D model. Figure 6 shows how the cell voltage is maximum at the origin of the x-y axis, corresponding to the inlet of both solutions, while it is minimum in the opposite corner, corresponding to the outlet of the two solutions. Also in this case, the main variation in cell voltage occurs along the diluate flow direction, though a fairly important variation is also observed along the concentrate flow direction.



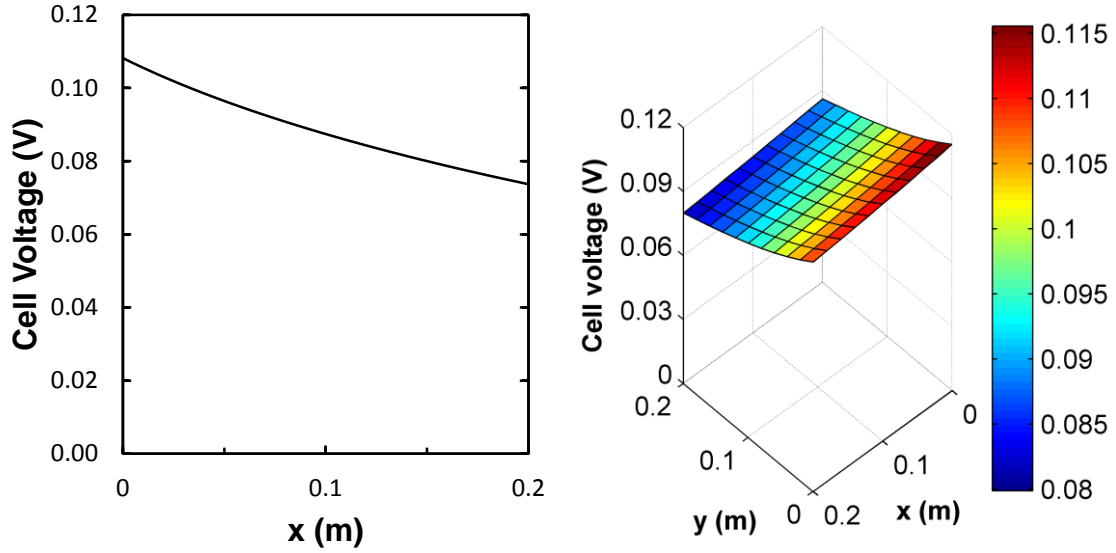


Figure 6. Voltage profile within a cell pair estimated by the previous 1-D co-current model (left, [17]) and by the 2-D cross-flow model (right). Simulations of a  $20 \times 20$  cm<sup>2</sup> 50 cell pairs stack equipped with Fujifilm membranes, 270  $\mu$ m woven spacers;  $C_{HIGH} = 5$  M,  $C_{LOW} = 0.1$  M; fluid velocity: 0.5 cm/s;  $T=20^{\circ}\text{C}$ .

### 3.2 Effect of different path lengths for diluate/concentrate streams

It has been already highlight that the effect of diluate and concentrate on the overall stack resistance is sensibly different [19]. In order to simulate different path length for diluate and concentrate streams, the aspect ratio of the equipment has been defined as:

$$Aspect\ Ratio(AR) = \frac{HIGH\ path\ length(L)}{LOW\ path\ length(b)} \quad (9)$$

According to such definition, an aspect ratio  $< 1$  means a short compartment for concentrate and longer for diluate. The effect of aspect ratio on process performance is shown in Figure 7 and Figure 8 for a 100 cell pairs stack. Simulations were carried out assuming alternatively a fixed path length of 20 cm for one of the two streams, and increasing the path length for the other up to 5 times (i.e. from 20 up to 100 cm). Flow rates of both solutions were adjusted accordingly, in order to keep the same fluid velocity (1 cm/s) inside all channels.

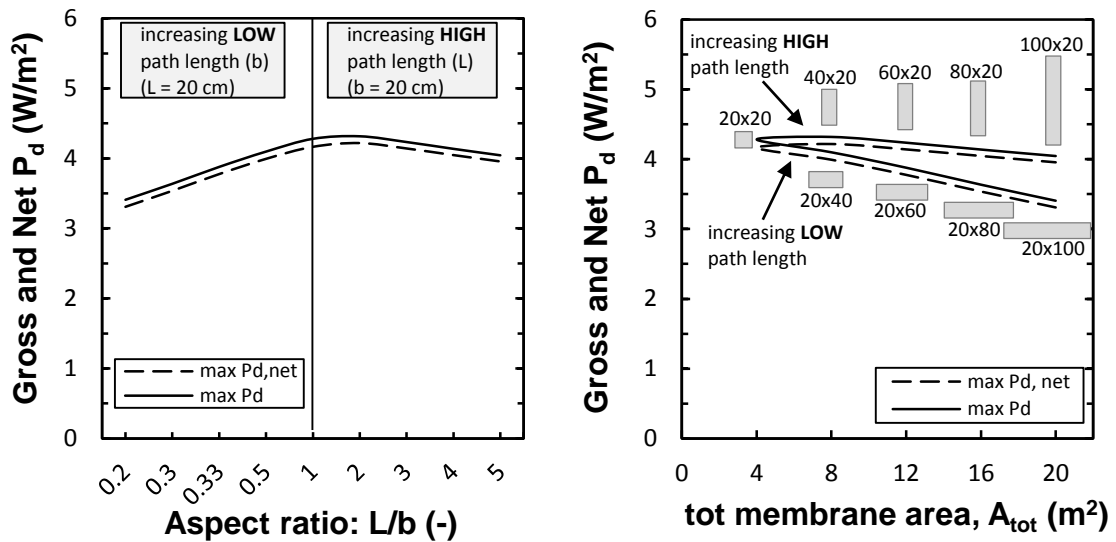


Figure 7. Influence of aspect ratio on gross and net power density. Simulations of a 100 cell pairs stack equipped with Fujifilm membranes, 270  $\mu m$  woven spacers;  $C_{LOW} = 0.1 M$ ,  $C_{HIGH} = 5 M$ ;  $v_{HIGH} = v_{LOW} = 1 cm/s$ ;  $T=20^\circ C$ . The same model predictions are shown as a function of aspect ratio (left plot) or total membrane area (right plot).

Figure 7 shows that the power density is negatively affected by a longer path length of the two streams, due to the increased residence time and consequent reduction in the average concentration gradient available. In particular, the increase in the diluate path length leads to a significant reduction in the power density generated. Conversely, a longer path length for concentrate has a weaker effect. In particular, increasing the paths length ratio from 1 (i.e. 20x20  $cm^2$  symmetric stack) up to 5 (20x100  $cm^2$  stack), a decrease in the power density of 20% and 7% are predicted by the model for the first and the second case, respectively. A local maximum in the power density is predicted for  $AR = 2$ , where the  $P_d$  is slightly higher than for symmetrical stack. Such initial increase is likely related to a beneficial effect of the reduced mean concentration of the concentrate stream on membranes permselectivity. On the other hand, the global reduction is clearly related to the average reduction of driving force within the stack. On this basis, increasing the concentrate path length does benefit more the power output in terms of total net power produced (Figure 8), indicating how the use of asymmetric stack, using a shorter path for diluate, allows to enhance the power output and yield (defined with respect to the HIGH feed solution) of a RED process.

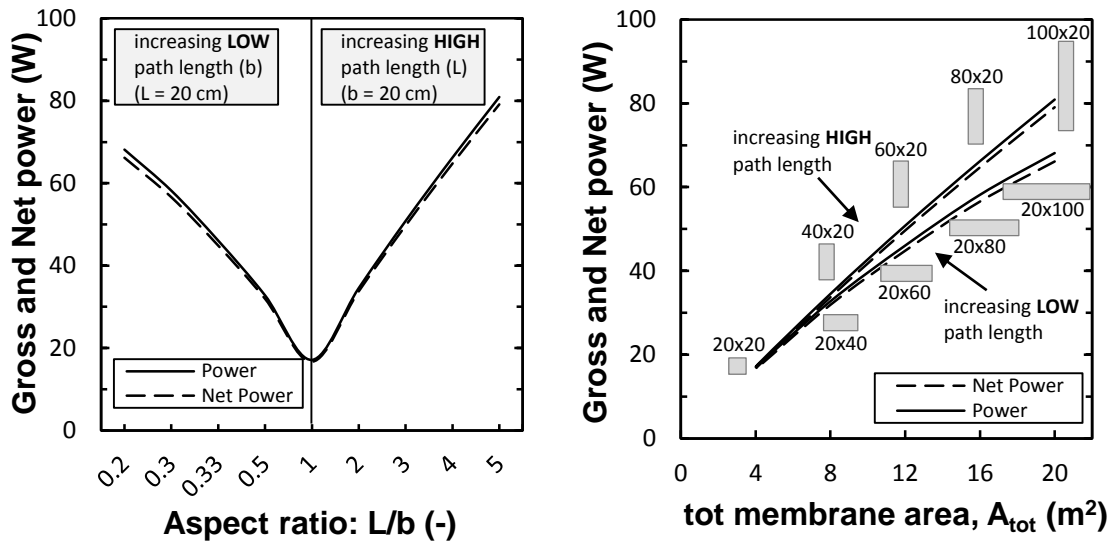


Figure 8. Influence of aspect ratio on gross and net power. Simulations of a 100 cell pairs stack equipped with Fujifilm membranes, 270  $\mu\text{m}$  woven spacers;  $C_{\text{LOW}} = 0.1 \text{ M}$ ,  $C_{\text{HIGH}} = 5 \text{ M}$ ;  $v_{\text{HIGH}} = v_{\text{LOW}} = 1 \text{ cm/s}$ ;  $T=20^\circ\text{C}$ . The same model predictions are shown as a function of aspect ratio (left plot) or total membrane area (right plot).

### 3.3 Exploring the optimal operating conditions

#### 3.3.1 Effect of salt concentration

The effect of inlet concentration for both diluate and concentrate has been investigated on a 20x20 cm<sup>2</sup> stack equipped with 100 cell pairs (Figure 9). It is worth noting that feed concentrations affect also the viscosity of solutions and, therefore, the pressure drops inside the stack. As a consequence, the effect of inlet concentration on net power density is slightly different (Figure 9b).

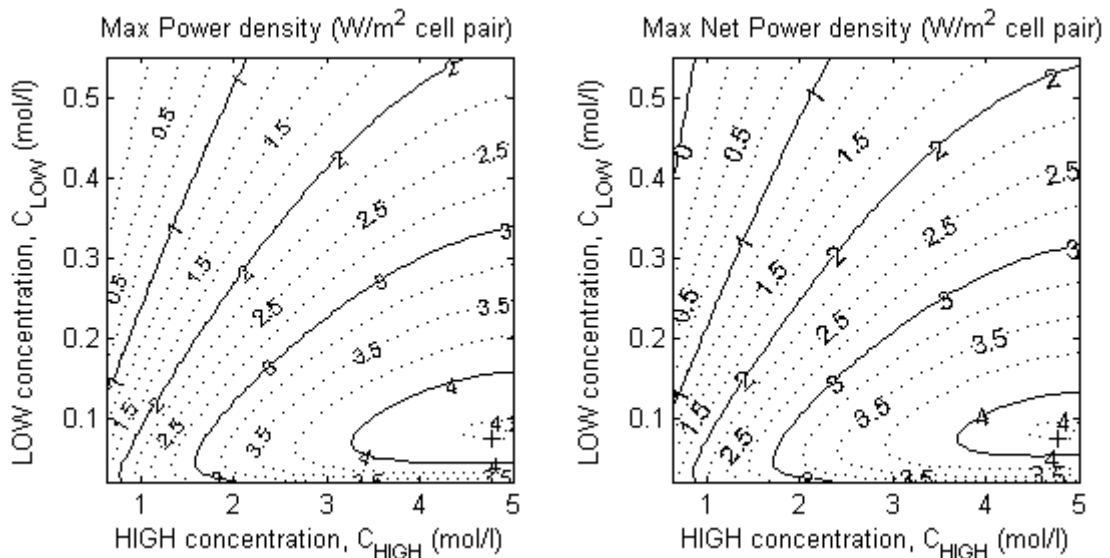


Figure 9. Effect of the inlet concentration on gross and net power density for a 20cmx20cm 100-cells stack, equipped with Fujifilm membranes, 270  $\mu\text{m}$  woven spacers. Feed flow velocity: 1 cm/s;  $T=20^\circ\text{C}$ .

Interestingly, the maximum power density is achieved by the 20x20 cm<sup>2</sup> stack at inlet concentration of approximately 0.08 M NaCl and 4.7 M NaCl, which can be identified as

optimal solutions for the investigated case. It is worth noting, that such conditions lie in a similar concentration range as the one previously identified by the same authors for a 10x10 cm<sup>2</sup> stack [20]. On the other side, the net power density map (Figure 9b) shows a zone where negative values are predicted, indicating that the pumping power is greater than the gross power output.

### 3.3.2 Influence of feed flow rates

Starting from the optimal feed concentration shown in Figure 9, different conditions were explored in order to find the best flow rates for both solutions. Figure 10 shows gross and net power density for a 20x20 cm<sup>2</sup> 100-cells stack assuming a fluid velocity ranging from 0.1 up to 3 cm/s for both solutions.

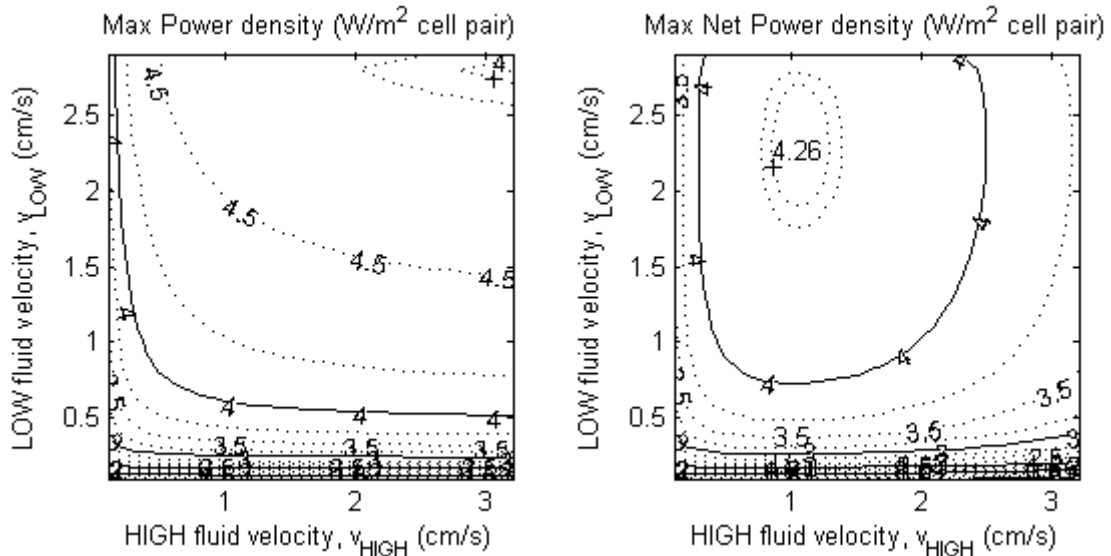


Figure 10. Effect of channel velocities on gross and net power density for a 20cm×20cm 100-cells stack, equipped with Fujifilm membranes, 270  $\mu$ m woven spacers;  $C_{LOW} = 0.08$  M,  $C_{HIGH} = 4.7$  M;  $T=20^{\circ}$ C.

Clearly, the higher the fluid velocity, the higher will be the gross power density (Figure 10a), due to the low residence time for both solutions inside the stack and relevant larger driving forces kept between the membranes. However, it is worth noting that the maximum for the diluate velocity is not achieved at the boundary. In fact,  $P_d$  increases until an optimal velocity of about 2.7 cm/s is reached, while for larger values of velocity  $P_d$  decreases again. This is related to the effect of the increase in electrical resistance due to the low conductivity of the solution, generated by extremely small residence times.

On the other side, some interesting results are shown in Figure 10b, where the net power density is plotted as a function of LOW and HIGH fluid velocity. Model predictions show that a maximum net power density of 4.6 W/m<sup>2</sup> can be reached choosing a fluid velocity of 1 cm/s for concentrate and between 2 – 2.5 cm/s for diluate. This allows to ensure high driving forces, but with acceptable pressure drops.

All the results shown so far indicate how the system performance is sensibly affected by diluate properties. In order to explore in more detail the optimal operating conditions for a 100-cells stack (20x20 cm<sup>2</sup>), a further analysis has been carried out changing both velocity and inlet concentration for the diluate. Results are shown in Figure 11, where gross and net power density are plotted as a function of the operating conditions for diluate, ranging from 0.1 to 3 cm/s for fluid velocity, and from 0.01 to 0.55 M for inlet concentration. All the

simulations were performed changing the concentrate fluid velocity accordingly with the diluate velocity.

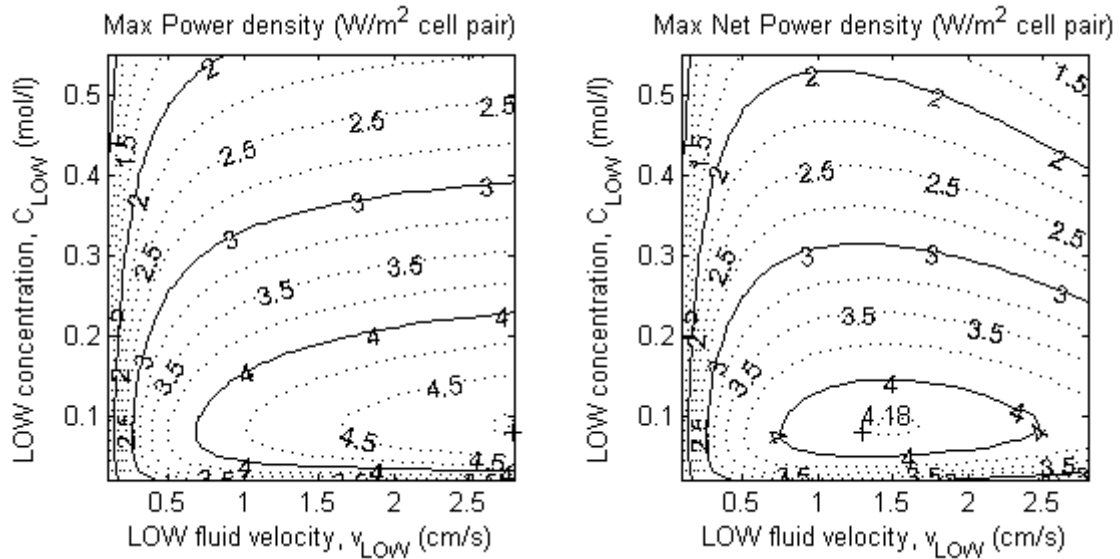


Figure 11. Gross and net power density as a function of operative conditions (inlet concentration in diluate compartment and fluid velocity). Simulations of a 20cm×20cm 100-cells stack, equipped with Fujifilm membranes, 270  $\mu\text{m}$  woven spacers;  $C_{\text{HIGH}} = 4.7 \text{ M}$ ;  $v_{\text{HIGH}} = v_{\text{LOW}}$ ;  $T=20^\circ\text{C}$ .

Given the assumption of equal fluid velocities for both streams, a maximum net power density of  $4.18 \text{ W/m}^2$  is predicted by the model, in conditions slightly different than in the previous cases, i.e. with a fluid velocity of  $1.5 \text{ cm/s}$  and a LOW inlet concentration close to  $0.1 \text{ M NaCl}$  (Figure 11b).

### 3.4 Simulation of a pilot-scale unit

In order to simulate a larger RED unit (with reference, in particular, to the prototype unit designed for the operation with saltworks brines in Trapani, Italy, within the REAPower project [21]), a number of different scenarios were taken into account starting from a  $22 \times 22 \text{ cm}^2$  stack (100 cells) as reference case (Table 1). All scenarios were designed to investigate the effect of scale-up on either symmetric or asymmetric stack design. A concentrated brine of  $5 \text{ M NaCl}$  and brackish water ( $0.1 \text{ M NaCl}$ ) were considered as feed solutions for all cases. When simulating asymmetric stacks, the concentrated brine was always fed to the longer channel, given the better performance registered in the previously reported analysis for such flow configuration.

Table 1. Summary of simulations carried out on large prototype unit.

<b>Scenario #</b>	<b>membrane size (cm)</b>	<b>N° cell pairs</b>	<b>Total membrane Area (m<sup>2</sup>)</b>	<b>Description</b>
1	22 x 22	100	4.8	Reference case (small prototype)
2	22 x 22	500	24.2	Larger number of cell pairs
3	44 x 44	500	96.8	Larger symmetric stack
4	22 x 88	500	96.8	Larger asymmetric stack, AR = 4
5	44 x 88	500	193.6	Larger asymmetric stack, AR = 2
6	22 x 88	500	96.8	asymmetric stack, different velocity (v <sub>LOW</sub> = 1 cm/s, v <sub>HIGH</sub> = 2 cm/s)

Figure 12 shows the gross and net power density for the investigated scenarios. As expected, the decrease in power density due to the longer HIGH path length is rather negligible. Moreover, increasing the number of cell pairs up to 500 (scenario 2), causes a slight reduction in power density, likely due to the following two counteracting effects: 1) reduction of the influence of blank resistance on the overall performance; 2) increase of stack parasitic currents reducing the power output.

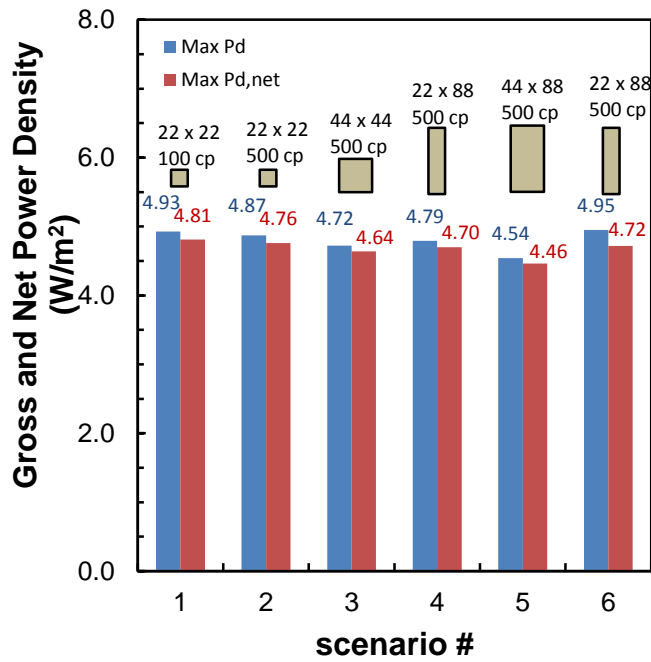


Figure 12. Gross and net power density for stacks equipped with Fujifilm membranes, 270  $\mu\text{m}$  woven spacers;  $C_{\text{LOW}} = 0.1 \text{ M}$ ;  $C_{\text{HIGH}} = 5 \text{ M}$ ;  $v_{\text{LOW}} = v_{\text{HIGH}} = 1 \text{ cm/s}$ , in scenario no. 6  $v_{\text{LOW}} = 2 \text{ cm/s}$ ,  $v_{\text{HIGH}} = 1 \text{ cm/s}$ ;  $T=30^\circ\text{C}$ .

Figure 13 shows the predicted gross and net power output for the above-mentioned scenarios. In the case of 44x88  $\text{cm}^2$  stack equipped with 500 cell pairs (scenario 5) a power output of almost 900 W has been predicted, which emphasize how the target of 1 kW can be reached with a single stack provided that an active membrane area of 44x88  $\text{cm}^2$  is guaranteed.

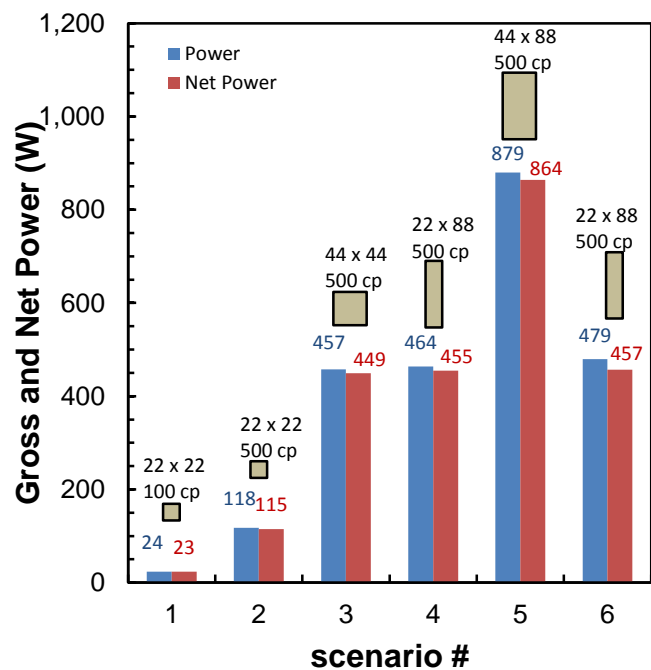


Figure 13. Gross and net power density for stacks equipped with Fujifilm membranes, 270  $\mu\text{m}$  woven spacers;  $C_{\text{LOW}} = 0.1 \text{ M}$ ;  $C_{\text{HIGH}} = 5 \text{ M}$ ;  $v_{\text{LOW}} = v_{\text{HIGH}} = 1 \text{ cm/s}$ , in scenario no. 6  $v_{\text{LOW}} = 2 \text{ cm/s}$ ,  $v_{\text{HIGH}} = 1 \text{ cm/s}$ ;  $T=30^\circ\text{C}$ .

### 3.5 Process simulation of a plant with 3 RED units

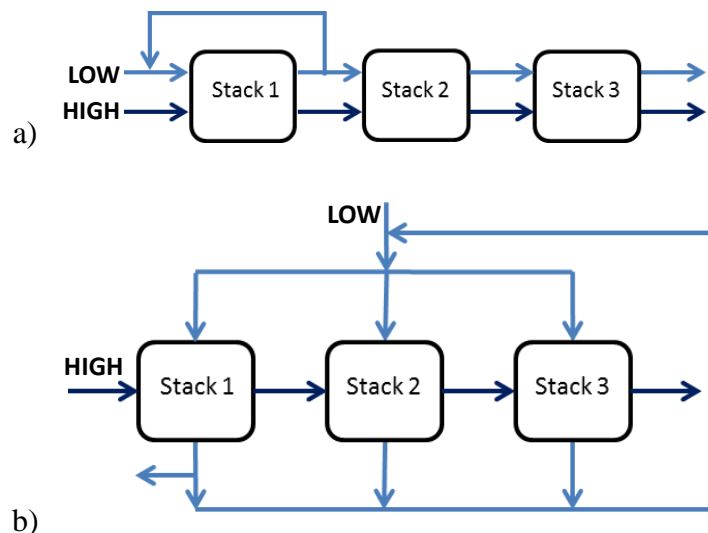
The developed process simulator was finally used to describe the operation of a whole RED plant with a train of 3 units, each one with a 44x44 cm<sup>2</sup> active membrane area and 500 cell pairs. The main goal of this modelling task was to simulate the possible plant layouts to be adopted for the REAPower demonstration plant [21], using concentrated brine and brackish water as feed solutions.

Aside from the stack design, the streams arrangement among the units can significantly affect the overall power output, as well as the performance of each single stack.

In all simulations the assumption was made of having an HIGH feed flow rate of 29.4 lt/min, in order to guarantee the standard internal velocity of 1 cm/s. However, three different scenarios in terms of diluate feed flow rates were investigated, to analyze the effect of different feed rates availability in the real installation site. In particular the flow rates of 20 lt/min, 29.4 lt/min (as in the HIGH channels) and 40 lt/min were set in the three simulation sets. Moreover, the dilute feed concentration was set to 0.03 M, according to the real availability of brackish water in the prototype installation site.

As previously shown, the residence time of concentrated brine has been found not to be dramatic for process performance. For this reason, a serial arrangement among units was adopted for the concentrate. Conversely, a number of possible distribution paths have been analyzed for the diluate. In Figure 14 the three most promising configurations are shown, namely:

- *Layout 1 (serial arrangement, Figure 14a)*. Three stacks are fed in series for both diluate and concentrate, with a recycle for the diluate in the first stage, allowing the control of fluid velocity of 1 cm/s inside all channels of the first stack, even if the availability of brackish water is low. This arrangement also guarantees in the first stack a more uniform concentration in the dilute compartment, in a range closer to the previously identified optimal concentration.
- *Layout 2 (parallel arrangement, Figure 14b)*. All stacks are fed in series for the concentrate stream, and in parallel for the diluate, thus requiring a partial recycle of the exiting dilute stream in order to operate all the three stacks with the standard velocity of 1 cm/s. This configuration ensures more homogeneous driving force for all units.
- *Layout 3 (parallel-serial arrangement, Figure 14c)*: this hybrid configuration provides a parallel feed for diluate into the first two stacks, using the overall outlet to feed the third unit, where the fluid velocity will be much larger than in standard conditions.





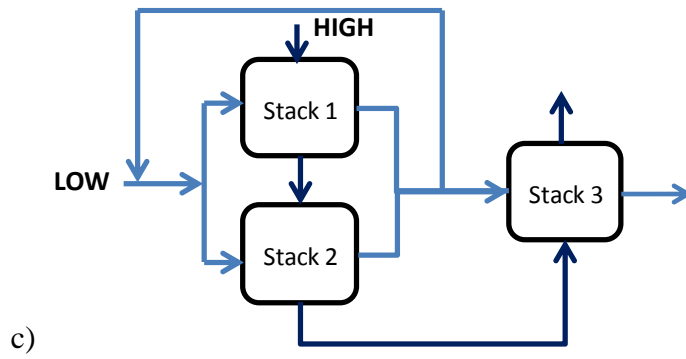


Figure 14. Investigated layouts for streams arrangements. a) Serial arrangement; b) parallel arrangement; c) parallel-serial arrangement.

The implemented flowsheet in gPROMS Model Builder environment [22] for the first layout is shown in Figure 15.

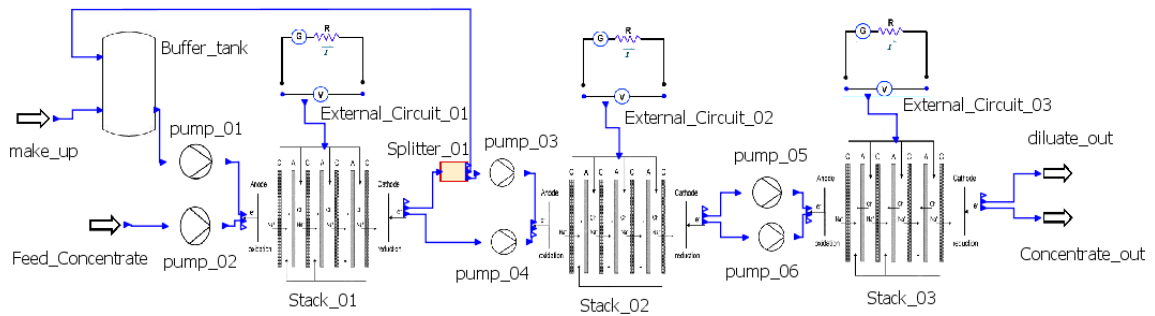


Figure 15. Implemented flowsheet for 3 RED units.

The first set of model predictions, relevant to the case of diluate feed flow rate of 20 l/min, are reported in Figure 16.

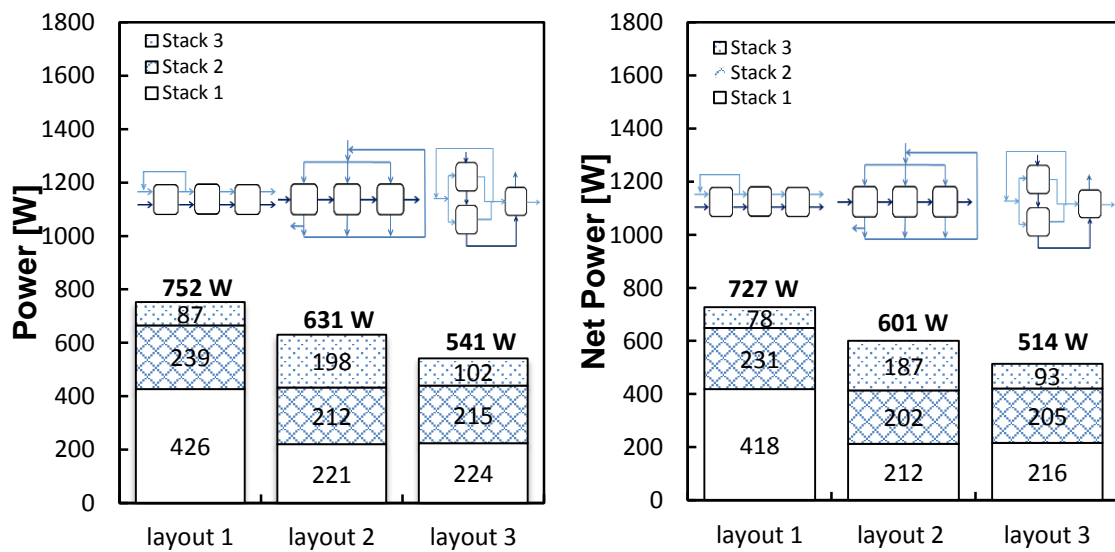


Figure 16. Gross and Net Power of 3 stacks (500 cells) equipped with Fujifilm membranes 44×44 cm and 270 μm woven spacers, in layout 1: serial; layout 2: parallel; layout 3: parallel-

serial for diluate feed;  $C_{HIGH} = 5M$ ;  $Q_{HIGH} = 29.4$  l/min; make-up of brackish water,  $Q_{MU} = 20$  l/min,  $C_{MU} = 0.03M$ .

For layout 1 (serial arrangement) a make-up of 20 l/min of brackish water is not enough to guarantee a velocity of 1 cm/s in all channels. For this reason, a recycle of the outlet diluate is required at least after the first stack. It is worth noting that layout 2 (parallel arrangement) is clearly the optimal configuration to ensure equal performance for all stacks. Nonetheless, layout 1 (serial arrangement) gives a higher power output (752 W), while the layouts 2 and 3 can provide (with 3 stacks) the same power output generated by the first 2 stacks in layout 1.

In order to feed the plant with no recycle and keep a fluid velocity of 1 cm/s in all channels, the required brackish water flow rate is 29.4 l/min. Figure 17 shows model predictions for the investigated plant layouts assuming the availability of 29.4 l/min of brackish water as make-up to the plant.

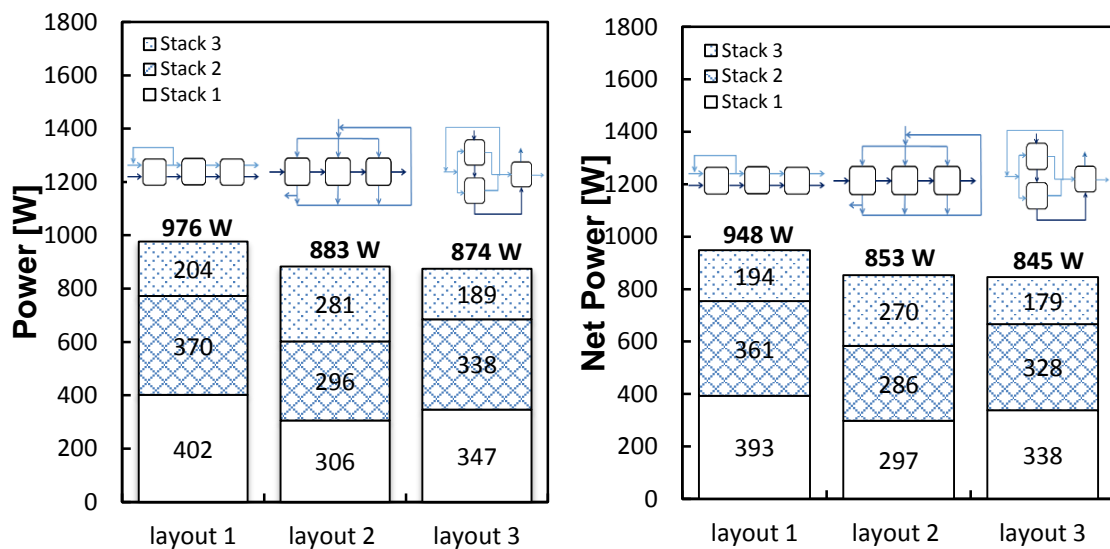


Figure 17. Gross and Net Power of 3 stacks (500 cells) equipped with Fujifilm membranes 44×44 cm and 270 μm woven spacers, in layout 1: serial; layout 2: parallel; layout 3: parallel-serial for diluate feed;  $C_{HIGH} = 5M$ ;  $Q_{HIGH} = 29.4$  l/min; make-up of brackish water,  $Q_{MU} = 29.4$  l/min,  $C_{MU} = 0.03M$ .

An increase of the diluate make-up allowed to reach a power output of 976 W for layout 1 (Figure 17). Moreover, a higher make-up of brackish water gives minor differences among the layouts in terms of power output, significantly improving the performances of layouts 2 and 3.

The availability of brackish water is a key factor for the plant performance. Assuming to double the original amount of brackish water available, the process simulator was eventually used to predict the plant operation with a make-up of brackish water of 40 l/min (Figure 18).

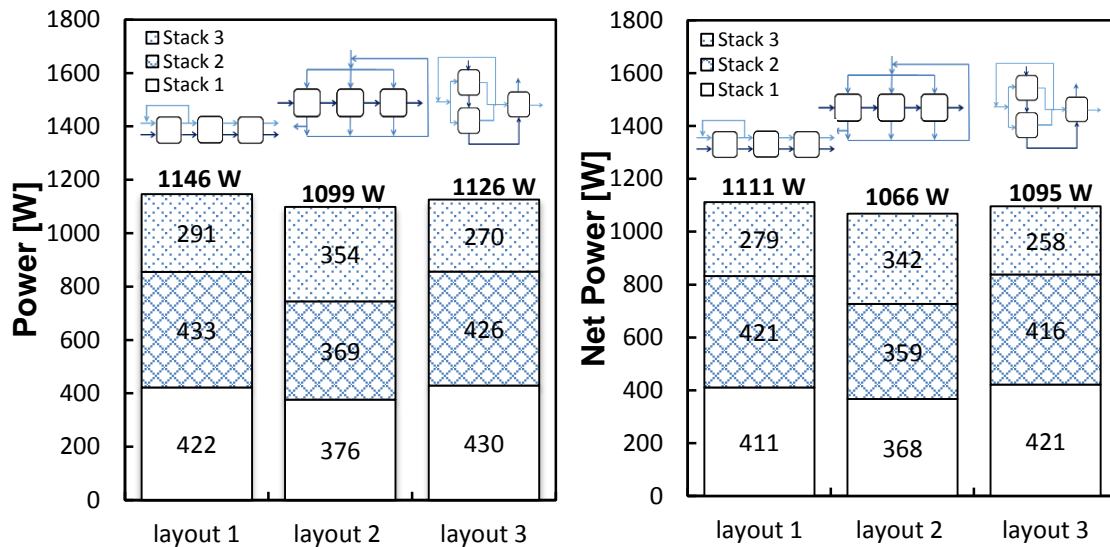


Figure 18. Gross and Net Power of 3 stacks (500 cells) equipped with Fujifilm membranes  $44 \times 44 \text{ cm}^2$  and  $270 \text{ }\mu\text{m}$  woven spacers, in layout 1: serial; layout 2: parallel; layout 3: parallel-serial for diluate feed;  $C_{\text{HIGH}} = 5\text{M}$ ;  $Q_{\text{HIGH}} = 29.4 \text{ l/min}$ ; make-up of brackish water,  $Q_{\text{MU}} = 40 \text{ l/min}$ ,  $C_{\text{MU}} = 0.03\text{M}$ .

A brackish water availability of  $40 \text{ l/min}$  would allow to reach more than  $1 \text{ kW}$  with all of the investigated plant layout. Interestingly, for layout 1 the power produced from the second stage ( $433 \text{ W}$ ) is slightly higher than the power from the first stage ( $422 \text{ W}$ ). This is because in the second stage the diluate inlet concentration is closer to its optimal value, as previously shown by the parametric analysis (Figure 10).

#### 4 Conclusions

This work presents a new process simulator for a Reverse Electrodialysis plant using sea/brackish water and concentrated brines as feed solutions. Starting from a previous work by the same authors, a new model was implemented for simulating the performance of a RED unit with cross-flow arrangement. Regarding the stack design, the possibility of using asymmetric stack has been addressed, being the model able to simulate different flow path lengths for dilute and concentrate streams, adopting a  $20 \times 20 \text{ cm}^2$  stack with 100 cell pairs as reference case. Model predictions showed how increasing the aspect ratio from 1 (i.e.  $20 \times 20 \text{ cm}^2$  symmetric stack) up to 5 ( $20 \times 100 \text{ cm}^2$  stack), a decrease from 7% to 20% in the power density is registered, with larger effects observed when increasing the diluate channel length. On the other side, increasing the concentrate path length enhances almost proportionally the power output and the overall process yield (defined with respect to the concentrate stream) achieved.

An optimisation analysis has been performed in order to identify optimal operating conditions for a  $20 \times 20 \text{ cm}^2$  stack in terms of optimal inlet feed concentration and velocities, highlighting the dominant influence of diluate compartment conditions on the overall unit performance.

Finally, simulations were performed to predict the behaviour of a larger prototype plant, being this a preliminary analysis for the construction of a demonstration plant to be installed in Trapani (Italy) as a final deliverable of the REAPower project [21]. A number of possible layouts have been presented, connecting 3 RED units equipped with 500 cells pairs of  $44 \times 44 \text{ cm}^2$  active membrane area. Three different conditions were considered in terms of availability of diluate feed. Interestingly, in the case of low availability of brackish

water (20 l/min), a serial arrangement is preferable, with 2 stacks in series providing practically the same power output as 3 stacks differently connected in the other two layouts. However, assuming an availability of brackish water of 40 l/min, simulations indicate a much closer performance between the three layouts, allowing to reach in all cases more than 1 kW power output.

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